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CHEMICAL CONSTRUCTION CORPORATION

DESIGN STUDY  
HYDROGEN GENERATION  
FROM LIGHT PETROLEUM DISTILLATE  
FOR  
BUREAU OF SHIPS CODE 660S

1/21/80

CHEMICAL CONSTRUCTION CORPORATION  
NEW YORK, NEW YORK

Job 1378  
September 28, 1962

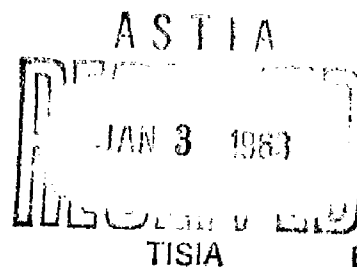


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I. INTRODUCTION AND SUMMARY

A. INTRODUCTION

A feasibility study was carried out by the ~~Chemical Construction Corporation~~ to determine if a practical hydrogen generator using the catalytic reforming of a light petroleum distillate, such as JP-5, could be designed. This hydrogen generator would be used to supply hydrogen for the generation of electricity in a hydrogen-oxygen fuel cell in submarine service. The study was conducted under Contract Nobs #86747, Project Serial No. SF 013-06-04, Task 4377.

The hydrogen generator must be capable of a minimum of ten days of continuous operation, without shutdown for maintenance or replacement of parts or components. The normal output of pure hydrogen must be 20 pounds per hour. The system must be capable of operating with an output range of from five pounds per hour at stand-by to a maximum of 76 pounds per hour. The maximum rate must be available for a total of ten hours operating time, in two hour periods.

The hydrogen generator must be optimized for maximum efficiency and reliability. It must minimize hazard and maintenance. Space requirements must be reduced to a practical minimum. The total hydrogen inventory must not exceed 900 standard cubic feet of hydrogen which is equivalent to 3% of the volume of the submarine.

The following items must be provided:

1. Heat and material balances
2. Process flow diagrams
3. Evaluation of startup and response times
4. Engineering assumptions which require further tests
5. Arrangement of equipment
6. Instrumentation

B. SUMMARY

The process proposed by ~~Chemico~~ for the production of hydrogen by the reforming of liquid hydrocarbons involves established and proven technology.

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In the Chemico process, vaporized hydrocarbon is catalytically reacted with steam and oxygen at elevated pressure and temperature. The resulting process gas stream consists of hydrogen, carbon dioxide and carbon monoxide, together with excess steam. The carbon monoxide in the gas stream is reacted with steam in a second catalyst bed, to form additional hydrogen plus carbon dioxide.

Pure hydrogen is separated from the final gas stream by diffusion. This method of separation employs a permeable membrane or partition composed of a palladium-silver alloy. The membrane selectively permits hydrogen to pass through, while holding back other gas components. The pure hydrogen is then collected from the other side of the membrane and is passed to the fuel cell for utilization.

After the separation of the hydrogen, the residual gas stream is catalytically reacted by controlled addition of oxygen to achieve complete combustion. The heat generated by this catalytic reaction is used to generate the steam required in the process. The gas stream, containing carbon dioxide and water vapor, is discharged overboard.

The hydrogen generator will operate in accordance with good engineering practice, employing process principles and equipment developed by Chemical Construction Corporation and tested either by the company or by suppliers.

Chemico has extensive and varied experience in the design of equipment for reforming and other high temperature and high pressure processes. In the past ten years Chemico has engineered over 40 plants incorporating more than 100 reform furnaces with a wide range of capacities. Much of the equipment, including the patented Chemico reform furnace, used in these plants was specially designed by Chemico for each particular application.

Engineering flow diagrams will be provided in Section IV Appendix together with heat and material balances. The present study will serve as a basis for a subsequent development study, which will include pilot plant testing of the hydrogen generator design.

### C. CONCLUSIONS

1. It is feasible to use the Chemico reform process to generate hydrogen in the quantities desired. The process requirements are as follows.

To generate one pound of hydrogen:

3.45# of  $C_{14}H_{30}$

4.18# of  $O_2$

2. This process has proven reliable in pilot plant tests on similar hydrocarbons. These tests have proven that proper selection of operating conditions eliminate the formation carbon within the catalyst bed.
3. The process offered is essentially noiseless and requires few moving parts. These parts consist of several small pumps which operate continuously and a compressor which operates only when the sea pressure exceeds the operating pressure of the process.
4. The process offered is versatile. It may be operated either automatically or manually. It may be operated at any rate within the design range for any period of time dependent only on the fuel, oxygen and caustic supplies available.
5. The rate of startup is dependent entirely on the rate at which heat can be brought into the system to raise it to operating temperature. With all equipment cold the maximum starting time is in the order of two hours.
6. Once the system is operating the response time for change in rate should be in the order of seconds.
7. The efficiency of the system is based on the ratio of the heat of the product hydrogen to the heat input of the fuel plus the energy requirements for the operation of the system. This is 77.5% at the maximum rate of 76 pounds per hour of hydrogen. The efficiency at other rates will be of the same magnitude since the input of the fuel and the energy requirements are dependent on the rate of generation of the hydrogen.
8. The storage requirements of the system are 0.068 cubic feet of fuel and 0.058 cubic feet of oxygen per pound of hydrogen produced.

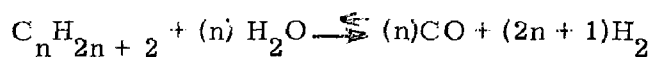
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9. The fuel used is logistically available and easy to store. It may be stored in the same manner as diesel fuel and is immiscible with sea water, minimizing the ballast problems which might be encountered. The fuel is inexpensive, reducing the overall operating costs.
10. Because of the fact that desulfurization equipment is included, it is possible that a single fuel may be used for hydrogen generation as well as for the diesel engine.
11. The Chemico system offered is compact. It can be fitted into the space allotted with a minimum of changes to existing equipment. The volume of hydrogen inventory is less than the 900 scf allowable, which is 3% of the volume of the submarine.
12. The system offered allows "scale-up" to larger size units with a minimum of difficulty.
13. The hydrogen produced is of ultra high purity and is well suited for fuel cell applications. In order to protect this purity the diffuser, which is essential to the production of high purity hydrogen, has been divided into four sections each of which can be isolated in the event of failure. This arrangement will allow servicing of any section without interruption of three fourths of the hydrogen output.
14. Carbon dioxide and water containing small quantities of sodium sulfide and sodium hydroxide will be the only waste materials discharged to the sea.
15. Reactors used in this system were specifically designed by Chemico for this application. The primary design considerations were Reliability, Safety and Minimum Volume and Response Time.

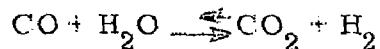
II. PROCESS AND EQUIPMENT DESIGNA. PROCESS DESCRIPTION

The process is designed to employ the liquid petroleum distillate J.P.-5 as a major process component. This material has a maximum sulfur content of 0.4%, however, the sulfur tolerance of the catalysts, diffuser and fuel cell is very low. Therefore, a continuous system of desulfurization has been provided. In the desulfurization process, J.P.-5 is heated at elevated pressure to a temperature of about 600°F and is mixed with hydrogen. The mixture is catalytically reacted to convert organic sulfur compounds to hydrogen sulfide. After cooling to 100°F, the process stream is washed with dilute aqueous caustic solution, to remove hydrogen sulfide. The gas phase and the aqueous and organic liquid phases are separated by decantation. The gas phase is passed to an oxidizer unit, where the hydrogen is catalytically reacted with oxygen to form water. The aqueous liquid phase, containing small quantities of sodium hydroxide and sodium sulfide, is discharged overboard. The organic liquid phase, consisting of desulfurized J.P.-5, is passed to process usage.

The desulfurized J.P.-5 is steam atomized to form a vapor stream which is reacted with oxygen and steam at a temperature of about 1,600°F. The J.P.-5 is thermally cracked to form hydrocarbons of lower molecular weight, such as methane, ethane, and ethylene. In addition, some combustion of the J.P.-5 also takes place, as well as reaction between steam and the J.P.-5 or other hydrocarbons. The resulting process stream now enters a reform catalyst bed at a temperature of about 1,900°F. Here the endothermic reform reaction between the naphtha or lower hydrocarbons and steam goes to completion.



The process gas stream leaving the reform unit contains carbon monoxide, carbon dioxide, hydrogen and steam. The gas stream is cooled from 1,700°F reform temperature to 960°F by heat exchange with incoming process streams, and is further cooled to 450°F by heat exchange with water whereby process steam is generated. Further steam is added to the process gas stream, and the combined gas stream is passed through a second catalyst bed in which the CO-OX or "shift" takes place. This reaction between carbon monoxide and steam yields further hydrogen and carbon dioxide.



The final process gas stream leaves the CO-OX unit at 650°F and is cooled to 200°F to reduce water content by condensation, increasing the partial pressure of hydrogen in the gas stream. After removal of condensed water, the gas stream is reheated to 600°F by heat exchange with the hot gas stream leaving the CO-OX unit. The gas stream is now passed to the diffuser unit, where the bulk of its hydrogen content is separated. The diffuser is provided with a permeable membrane composed of palladium-silver alloy, which selectively permits only hydrogen molecules to pass through, as pure hydrogen product. The product hydrogen is collected at low pressure from the other side of the membrane, and is cooled and delivered to the fuel cell for final usage.

The balance of the gas stream, consisting mainly of carbon dioxide and water vapor together with residual hydrogen, carbon monoxide and methane, leaves the diffuser at elevated pressure and passes to the oxidizer unit. Here the combustible gases are eliminated by controlled catalytic combustion, with addition of oxygen in successive stages. Generated heat is removed by simultaneous heat exchange with pressurized water, to provide a stream of saturated process steam.

The gas stream leaving the oxidizer principally contains carbon dioxide and water vapor, together with a trace of oxygen. This stream is cooled to recover water by condensation, and the residual gas stream is discharged overboard. A gas compressor is provided to pressurize the gas stream for discharge, when outside sea water pressure is equal to or exceeds the system operating pressure.

B. EQUIPMENT LISTE            -            Heat Exchangers and Condensers

26.1 E-1	JP-5 Preheater	12
26.2 E-1	Diffuser Feed Preheater.....	18
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H            -            Reactors

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26.3 H-1	O.S.C. (Organic Sulfur Converter)	9

J            -            Pumps

26.1 J-1	JP-5 Metering Pump	11
26.2 J-1	Boiler Feed Water Pump.....	23
26.3 J-1	JP-5 Feed Pump	8
26.3 J-2	Aqueous Phase Pump.....	10

K            -            Compressors

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26.1 R-2	Superheating Coil	)	13 thru 17
26.1 R-3	Water Heating Coil	)	
26.1 R-4	CO Converter	)	
26.2 R-1	Hydrogen Diffuser		20

C. DESCRIPTION OF EQUIPMENT

The actual location of the equipment and its relation to existing equipment in the submarine is shown in the photographs of the model. Copies of the flow diagrams and photographs are included in Section IV. The drawings are printed so that when unfolded they extend sufficiently beyond these pages to be referred to while reading the text.

1. DESULFURIZATION (Section 26.3)JP-5 Feed Pump (26.3 J-1)

Positive displacement gear type pump

Fluid Pumped: JP-5  
Capacity: 1 GPM  
Suction: Flooded  
Discharge Head: 200 psig  
Horsepower: 2 HP

O.S.C. Feed Preheater (26.3 E-1)

Heliflow type exchanger consisting of a single, spiral shaped coil held between two flat surfaces and enclosed in a welded shell. The coil forms a closed spiral shaped path for the shell fluid, for true counterflow heat exchange.

## Overall Dimensions:

Circular shell - 20" O.D. x 15" high

## Operating Conditions:

	<u>SHELL SIDE</u>	<u>TUBE SIDE</u>
Fluid	JP-5 Liquid	JP-5 Liquid + H <sub>2</sub> S + H <sub>2</sub>
Temp. in °F	77	600
Temp. out °F	570	100
Pressure psig max.	250	250

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Heat Exchange	-	95,700 BTU/Hr
Surface Per Unit	-	46 Ft <sup>2</sup>
MTD (Corrected)	-	30 °F
Transfer Rate Service	-	70 $\frac{\text{BTU}}{\text{Hr-Ft}^2 \text{ °F}}$

O. S. C. Vessel (Organic Sulfur Converter) 26.3 H-1 &  
26.3 E-2

In the O. S. C. vessel the organic sulfur compounds in the JP-5 are catalytically converted to H<sub>2</sub>S. The unit consists of a cylindrical steel shell containing a bed of cobalt - moly catalyst supported on steel grates. The support grates are covered with steel wire screen and stainless steel type 304 wire cloth.

An electric heating element is inserted in an open pipe at one end and flanged to the bottom of the shell at the other, forming an in line heater.

Heated JP-5 and H<sub>2</sub> enter the bottom of the O. S. C. vessel where additional heat is added by the electric heater to bring the mixture up to reaction temperatures. This mixture passes through the catalyst bed where the H<sub>2</sub> reacts with the sulfur compounds in the JP-5 to form H<sub>2</sub>S. The JP-5, H<sub>2</sub> and H<sub>2</sub>S leave the top of the O. S. C. and enter the O. S. C. feed heater.

### Overall Dimensions:

18" O.D. x 5' High

### Operating Conditions:

Pressure	-	250 psig
Temperature	-	600 °F

Phase Separator (26.3 G-1)

The function of this unit is to separate the JP-5 from the aqueous caustic layer containing  $\text{Na}_2\text{S}$ .

The JP-5, caustic solution, and  $\text{H}_2$  gas enter the bottom of the vessel. The caustic solution containing dissolved  $\text{Na}_2\text{S}$  settles at the bottom, while the JP-5 overflows to a storage chamber. Further settling of the two phases is accomplished in the storage chamber. JP-5 is drawn from the top of this storage chamber, while the caustic phase settles and drains to the bottom.

Hydrogen gas exits the top of the unit through a mist entrainment element.

Overall Dimensions:

18" O.D x 6' - 0" high

Operating Conditions:

Pressure	-	250 psig
Temperature	-	100 °F

Aqueous Phase Pump (26.3 J-2)

Two stage turbine type pump.

Fluid Pumped: 1% Caustic Solution

Capacity: 1 GPM

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Suction Pressure: 200 psig  
Max. Discharge Pressure: 250 psig  
Horsepower: 2 HP

Caustic Storage (26.3 G-2)

A vertical, cylindrical, steel tank which is filled with 50% caustic solution. The caustic is pumped from the tank by means of gas pressure.

Overall Dimensions:

30" O.D. x 42" high

Operating Conditions:

Temperature - 77°F  
Pressure - 250 psig max.

2. REFORMING (Section 26.1)

JP-5 Metering Pump (26.1 J-1)

Reciprocating Positive Displacement Duplex Pump. Stroke length and speed are automatically controlled.

The plungers on the two pumps operate 180° out of phase.

Discharge is through a four-step check valve.

Fluid Pumped: Desulfurized JP-5

Capacity: 1 GPM

Suction Pressure: 200 psig

Discharge Head: 250 psig

Horsepower: 2 HP

JP-5 Preheater (26.1 E-1)

Heliflow type exchanger consisting of a single, spiral shaped coil held between two flat surfaces and enclosed in a welded shell. The coil forms a closed spiral shaped path for the shell fluid, so that true counterflow heat exchange is achieved.

Overall Dimensions:

Circular shell - 13" O.D. x 10" High

Operating Conditions:

	<u>SHELL SIDE</u>	<u>TUBE SIDE</u>
Fluid	JP-5 Liquid	Hydrogen
Temp. in °F	100	600
Temp. out °F	500	300
Pressure psig max.	250	5
Heat Exchanged	-	81,000 BTU/Hr
Surface Per Unit	-	12 Ft <sup>2</sup>
MTD (Corrected)	-	130 °F
Transfer Rate Service	-	52 $\frac{\text{BTU}}{\text{Hr-Ft}^2 \text{ °F}}$

Integrated Reformer Unit (26.1 R1 thru R4)

a. The integrated reformer is designed to perform the following operations in one enclosed unit:

- 1) Burn the fuel with oxygen and mix the product with superheated steam.
- 2) Catalytically reform the mixed gas.
- 3) Catalytically convert the CO contained in the reformed gas to  $\text{CO}_2$ .
- 4) Preheat the feed water.
- 5) Superheat the steam.
- 6) Produce necessary heat for startup.

b. Reforming Section 26.1 R-1

- 1) The reformer is a cylindrical vessel, made of stainless steel, 24" O.D. and 65" high, internally lined with 3" thick refractory.

The upper end is closed by a flat, refractory lined cover. In the center of the cover is a special burner with nozzles for connection to fuel, oxygen and atomizing steam lines. The cover is bolted to a seal ring located on the top head of the unit. The cover is also bolted to the stainless reformer chamber.

The bottom end is formed by a stainless dished head. In the center of it is an opening protected by a perforated sleeve.

The center space is filled with granular catalyst forming a bed about 25" high. The top of the catalyst bed is covered by a 3" thick broken refractory protective layer. In the upper part of the cylinder, close to the burner, is located a superheated steam distributing sparger. The sparger is connected to the superheating coil by a pipe running inside the reformer and penetrating outside through an opening in the bottom head.

2) Heat Exchange Section 26.1 R-2, R-3

- a) The water preheater coil 26.1 R-2 consists of inlet and outlet straight tube headers, connected to a system of circular spiral wound coils made of extended surface carbon steel tubes.

The inlet and outlet headers are connected to external piping lines by internal connecting pipes and nozzles located on the top head of the unit.

- b) The steam superheater coil 26.1 R-3 is similar to the water coil but stainless steel is used for parts subjected to high temperatures.

The inlet nozzle is located on the top head and the outlet header is connected to the sparger inside the reformer vessel.

- c) Quench water sparger is a circular pipe ring located close to CO-OX section inlet. It is connected to external piping by a nozzle located at the top head of the unit.

- d) Internal basket is a thin stainless steel shell with lower dished head, internally insulated. It is open on the top and is supported by a perforated skirt.

3) CO Converter Section 26.1 R-4

The CO converter section 26.1 R-4 consists of catalyst filled annular space between the external vessel and internal basket.

The external vessel is of carbon steel, 72" O.D. and about 7 feet high, externally insulated and supported on four legs. The vessel is closed by a carbon steel dished cover connected to the vessel by bolted flanges. A dished head with the process gas outlet nozzle and drain connection is welded at the bottom.

The enclosing vessel is built in accordance with the ASME Code for unfired pressure vessels for pressures and temperatures required by process.

c. Operation

1) Process Gas Flow

In the burner the fuel is atomized by action of injected steam. It then comes in contact with inflowing oxygen and upon burning mixes with superheated steam, fed by the steam sparger.

The combined gases flow downward and pass through a broken refractory layer which acts as a distributor into the reform catalyst bed. After reforming the gas leaves through a perforated sleeve in the bottom of the reform vessel R-1. A circular space between the bottom of the reformer and heat exchanger basket serves as a gas collector and provides for even gas distribution into the heat exchange section. This section contains a steam superheating coil R-3 and a water preheating coil R-2. Upon passing the coils, the gases exchange heat with steam and water and cool down to the temperature level required for catalytic CO-Conversion.

Final temperature adjustment is made by injecting quench water which is vaporized in the gas stream. A sparger located at the inlet to the CO-Converter section is provided for this purpose.

From the heat exchange section the gases flow to the CO-Conversion section, R-4. The gases enter through a distributing layer of broken refractory above the catalyst bed. The catalyst bed is contained in the circular space between the outer shell and the heat exchange basket.

After the shift reaction has taken place the gas leaves through a perforated skirt, which serves also as a support for the heat exchange basket. The gas collects in the circular space provided between the basket and the outer shell, and exits through a nozzle in the bottom of the outer shell.

2) Water Flow R-2

The water is pumped at about 170° through the inlet header, flows through the heat exchange coil and leaves at 400°F. The stream, vaporized at this point to about 20%, proceeds to the oxidizer.

3) Steam Flow R-3

From the oxidizer, the saturated steam flows back through the inlet header and heat exchanger coil where by heat exchange countercurrent to outflowing reformed gas, it is superheated to 1625°F. Next the superheated steam flows through the outlet header, connecting pipe inside the reformer, and the outlet sparger. Here it is equally distributed around the burner for integral mixing with combustion gases.

d. Catalyst Handling

1) Reformer Catalyst

The reformer catalyst is loaded and unloaded by opening the flat reformer cover.

2) CO-OX Catalyst

The CO-Ox catalyst is loaded by opening the main top cover. For unloading two hand holes at the lower part of the shell are provided.

e. Thermal Stresses

The design provides for unrestricted independent expansion of all elements.

The thermal shock effects are eliminated by interposing wherever required a layer of insulated stainless steel.

The use of carbon steel for the pressurized external vessel was made possible by interposing a flow of cooled gases between the high temperature sections and the outside shell.

f. Maintenance

The access to the internals of the unit is provided by two covers. Every section is easily removable for inspection or maintenance.

The possibility to remove the reformer located in the center as a unit gives quick access to the heat exchangers and internals.

3. PURIFICATION (Section 26.2)Diffuser Feed Preheater (26.2 E-1)

Heliflow type exchanger consisting of a single, spiral shaped coil held between two flat surfaces and enclosed in a welded shell. The coil is surrounded by a plate baffle to prevent leakage of the shell fluid. The coil forms a closed spiral shaped path for the shell fluid, for true counterflow heat exchange.

## Overall Dimensions:

Circular shell - 24" O.D. x 16" High

## Operating Conditions:

	<u>SHELL SIDE</u>	<u>TUBE SIDE</u>
Fluid	Converter Gas & Steam	Saturated Converter Gas
Temp. in °F	650	200
Temp. out °F	375	600
Pressure psig max.	250	250
Heat Exchanged	-	227,000 BTU/Hr
Surface Per Unit	-	88 Ft <sup>2</sup>
MTD (Corrected)	-	100 °F
Transfer Rate Service	-	26 $\frac{\text{BTU}}{\text{Hr-Ft}^2 \text{ °F.}}$

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## Converter Gas Cooler (26.2 E-2)

Heliflow type condenser consisting of a single, spiral shaped coil held between two flat surfaces and enclosed in a welded shell. The coil forms a closed spiral shaped path for the shell fluid, for true counterflow heat exchange.

### Overall Dimensions:

Circular shell - 24" O.D. x 16" High

### Operating Conditions:

	<u>SHELL SIDE</u>	<u>TUBE SIDE</u>
Fluid	Cooling Water	Converter Gas & Condensate
Temp. in °F	95	375
Temp. out °F	120	200
Pressure psig max.	250	250
Heat Exchanged	-	566,600 BTU/Hr
Surface Per Unit	-	85 Ft <sup>2</sup>
MTD (Corrected)	-	170 °F
Transfer Rate Service	-	39 $\frac{\text{BTU}}{\text{Hr-Ft}^2 \text{ °F.}}$

Converter Gas Separator (26.2 G-2)

Line separator to remove entrained liquid from the converted gas. Unit is a vertical steel shell with S. S. internal separating element.

Overall Dimensions: 6-1/2" O.D. x 3' High

Hydrogen Diffuser (26.2 R-1)

The hydrogen diffuser separates hydrogen from the converter gas which contains  $H_2$ , CO,  $CH_4$ ,  $CO_2$ .

The unit consists of four assemblies, each one containing 25 diffusion cells. Each assembly is separately manifolded for feed gas, bleed gas, and product hydrogen gas. This arrangement allows one or more assemblies to be taken off stream in case of tube failure, without interrupting the process.

A single diffusion cell contains a large number of palladium-silver alloy tubes. Each tube is 1/16" O.D. x 0.003" wall thickness x 26" long. The tubes are enclosed in a stainless steel shell 1-3/4" O.D. x 28" long. The tubes are closed at one end and joined into a tube sheet at the other end. This design allows for a maximum transfer area with no restriction for thermal expansion, forming an ideal membrane

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Feed gas enters at the top of the shell and flows downward. The hydrogen in the gas diffuses from the outside to the inside of the palladium-silver tubes. The diffused hydrogen flows upwards and out through the hydrogen product manifold. The rest of the gas (bleed gas) leaves the cell through an axial located bleed gas collector tube.

Overall Dimensions: (Per Assembly)

4' Long x 4' High x 1' Wide

Operating Conditions:

### Pressure

Feed Gas - 200 - 250 psig

Bleed Gas - 200 - 250 psig

Hydrogen - 0 - 10 psig

Temperature - 600 - 800°F max.

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Catalytic Oxidizer (26.2 H-1)

The catalytic oxidizer has two functions:  
To oxidize the diffuser off gas and to generate steam for the process.

The vertical shell with finned fire tubes and dry drum-type boiler is used. The tubes are filled with catalyst and flanged for catalyst access. A 3/4" O.D. tube that runs concentrically in the catalyst bed is used for oxygen sparging. All the tubes are mounted on one flange and form one removable bundle. Concentric baffling is provided in the shell to guard against water splash.

Overall Dimensions:

Shell: 24" O.D. x 9' high

Tubes: 2" Finned tube, 3" O.D., 5' and 6' high

Drum: 12" O.D. x 2'-0" long

Operation:

Boiler feed water enters the vertical shell at the bottom. Steam is liberated at the top and exits through a baffle to a dry drum. Liquid from the dry drum drains back to the water shell.

Bleed gas from the diffuser containing 16.21% of combustibles is passed over the catalyst in the tubes. Hydrogen, carbon monoxide and methane are successively oxidized as the gas passes from tube to tube. Oxygen feed to each tube is proportioned for equal heat load distribution and temperature control. Exit gas is analyzed for residual oxygen. The analyzer controls a flow control valve in the oxygen heater.

CHEMICAL CONSTRUCTION CORPORATION

Boiler Feed Water Pump (26.2 J-1)

Reciprocating double acting piston pump.  
One standby pump provided.  
Both pumps are mounted on one base.

Fluid Pumped:	Water at 170 <sup>o</sup> F
Capacity:	3 GPM
Suction:	Flooded
Discharge Head:	250 psig
Horsepower:	3 HP

Condensate Collection Tank (26.2 G-1)

Process condensate and make up feed-water are stored in an "L" shaped rectangular vessel.

The base of the tank serves as a base for the feedwater pump.

Overall Dimensions:

Vertical Section: 24" High x 30" Wide x 6" Long

Horizontal Section: 3" High x 30" Wide x 21" Long

CHEMICAL CONSTRUCTION CORPORATION

Oxidizer Off Gas Cooler (26.2 E-3)

Heliflow type condenser consisting of a single, spiral shaped coil held between two flat surfaces and enclosed in a welded shell. The coil forms a closed spiral shaped path for the shell fluid, for true counterflow heat exchange.

Overall Dimensions:

Circular shell - 20" O.D. x 15" High

Operating Conditions:

	<u>SHELL SIDE</u>	<u>TUBE SIDE</u>
Fluid	Water	CO <sub>2</sub> and Condensate
Temp. in	95	500
Temp. out	120	120
Pressure psig max.	250	250
Heat Exchanged	-	204,000 BTU/hr
Surface Per Unit	-	54 Ft <sup>2</sup>
MTD (Corrected)	-	90°F
Transfer Rate Service	-	42 BTU Hr. - Ft. <sup>2</sup> °F

CHEMICAL CONSTRUCTION CORPORATION

CO<sub>2</sub> Gas Separator (26.2 G-3)

Line separator to remove entrained liquid from the CO<sub>2</sub> Gas. Unit is a vertical steel shell with S. S. internal separating element.

Overall Dimensions: 6-1/2" O.D. x 3' High

CO<sub>2</sub> Compressor (26.2 K-1)

The function of the compressor is to compress CO<sub>2</sub> from the system exit pressure of 200 psig to the pressure of the ambient sea water.

The compressor consists of two, single stage cylinders, with double acting pistons. Suction valve unloaders are included. The unit is driven by a D. C. motor mounted on the compressor crank case.

Overall Dimensions:

50" High x 41" Wide x 31" Long

Operating Conditions:

Suction Pressure	-	200 psig
Discharge "	-	500 psig max.
Capacity	-	124 SCFM
Discharge Temperature	-	250 °F max.
Brake HP @ 500 psig:	-	12 HP

CO<sub>2</sub> After Cooler (26.2 E-4)

Heliflow type exchanger consisting of a single, spiral shaped coil held between two flat surfaces and enclosed in a welded shell. The coil forms a closed spiral shaped path in the shell fluid, for true counterflow heat exchange.

## Overall Dimensions:

Circular shell - 10" O.D. x 9" High

## Operating Conditions:

	<u>SHELL SIDE</u>	<u>TUBE SIDE</u>
Fluid	Water	CO <sub>2</sub>
Temp. in °F	95	215
Temp. out °F	120	105
Pressure psig max.	250	500
Heat Exchanged	- 19,800 BTU/Hr	
Surface Per Unit	- 14 Ft <sup>2</sup>	
MTD (Corrected)	- 40 °F	
Transfer Rate	- 35 $\frac{\text{BTU}}{\text{Hr} \cdot \text{Ft}^2 \cdot ^\circ\text{F}}$	

CHEMICAL CONSTRUCTION CORPORATION

Hydrogen Gas Cooler (26.2 E-5)

Heliflow type exchanger consisting of a single, spiral shaped coil held between two flat surfaces and enclosed in a welded shell. The coil forms a closed spiral shaped path for the shell fluid, for true counterflow heat exchange.

Overall Dimensions:

Circular shell - 11" O.D. x 6" High

Operating Conditions:

	<u>SHELL SIDE</u>	<u>TUBE SIZE</u>
Fluid	Water	Hydrogen
Temp. in °F	95	300
Temp. out °F	120	140
Pressure psig max.	250	10
Heat Exchanged	- 44,330 BTU/Hr	
Surface Per Unit	- 4.2 Ft <sup>2</sup>	
MTD (Corrected)	- 105 °F	
Transfer Rate	- 100 $\frac{\text{BTU}}{\text{Hr} \cdot \text{Ft}^2 \cdot \text{°F}}$	

D. PROCESS CONTROL DESCRIPTION

The various controls referred to in this section can be identified on the engineering flow diagram. (Appendix, Section IV.)

Rate Control:

The hydrogen generation rate is from 5#/hr to 75#/hr. The rate of hydrogen generation is set by the JP-5 flow, which in turn sets the flow of steam and oxygen by ratio control.

JP-5 Desulfurizing Flow Control:

Liquid JP-5 is pumped through the heat exchanger and O.S.C. vessel into the phase separator. The upper section of the phase separator serves as storage for desulfurized JP-5. Level control LC-8 operating in the storage section regulates liquid control valve LCV-8 on the JP-5 pump discharge line. Level alarms indicate high and low levels. A level gage is provided to give a visual indication of the liquid level. Interface control IC-9 activates aqueous phase discharge valve LCV-9.

JP-5 Feed Control:

Desulfurized liquid JP-5 is fed by a piston type metering pump. The flow is controlled by piston speed and stroke adjustment from flow controller FRC-14. A signal from the JP-5 flow transmitter passes through a ratio controller and produces the correct set points on steam and oxygen flow recorder controllers, FRC-2 and FRC-11.

Oxidizer and Steam Generation Control:

The heat release in the oxidizer is controlled by bypassing converted gas, rich in hydrogen, around the hydrogen diffuser, thus increasing the combustibles in the oxidizer feed. Flow controller FRC-2 positioned in the boiler exit header cascades flow controller FRC-1. Flow control valve FCV-1, located in the diffuser bypass line, is controlled from flow controller FRC-1.

Boiler feed water flow into the oxidizer is regulated by the amount of bypass through flow control valve LCV-3, located on the discharge side of the feed water pump. LCV-3 is controlled by level controller LC-3, mounted on the oxidizer shell. High level and low level are indicated by level alarms. A level gage is provided to give a visual indication of the water level.

The oxidized gas exiting the oxidizer is analyzed for residual oxygen. The signal from the analyzer activates oxygen concentration controller ARC-6 and controls the oxygen flow to the oxidizer by means of flow control valve ACV-6.

CO<sub>2</sub> Discharge Pressure Control:

The CO<sub>2</sub> discharge pressure is set by pressure controller PRC-15 and pressure control valve PCV-15. If the system discharge pressure is higher than ambient sea water pressure, the CO<sub>2</sub> is discharged through pressure regulator VPC-16. If the system discharge pressure is lower than the ambient sea water pressure, VPC-16 closes. The compressor suction unloader is withdrawn and the gas is compressed to the ambient sea water pressure.

### III. DISCUSSION

#### A. SULFUR REMOVAL

The problem of desulfurization of the liquid petroleum distillate JP-5 could be solved by several alternative procedures. These included:

- a) purchase of sulfur-free material
- b) desulfurization of JP-5 as it is being loaded on the submarine
- c) continuous desulfurization of JP-5 immediately prior to usage.

Alternate a) was immediately eliminated, since this would limit the source of supply. Alternate b) was also eliminated, since the supply points would be limited to fixed desulfurization installations. In addition, alternate b) would require large-scale equipment because of the low space velocity required for liquid phase desulfurization. Thus for example, it was calculated that a catalyst bed having a volume of 130 cubic feet would be required for alternate b) assuming a 5-hour loading time. Therefore alternate c) namely continuous desulfurization of the JP-5 in the submarine immediately prior to usage, was selected for actual process design.

Continuous desulfurization may be carried out by several methods, such as sodium treatment, use of solid adsorbents, or liquid phase catalytic conversion. Such considerations as operating hazard and volume limitations formed the basis for rejection of all desulfurization methods except for liquid phase catalytic conversion.

The removal of sulfur from JP-5 by liquid phase catalytic conversion involves conventional technology. The process as described in this report is based on the Esso "Hydrofining" process which is similar to Universal Oil's "Unifining" process. In this process, the sulfur-containing hydrocarbon is mixed with free hydrogen, and the mixed stream is passed in contact with a cobalt-molybdenum catalyst at elevated pressure and temperature. The organic sulfur compounds are

converted to hydrogen sulfide, which is subsequently removed by a conventional method such as distillation or caustic wash. In the proposed process, a caustic wash will be employed. The chemical reaction between caustic and hydrogen sulfide is very rapid, consequently sufficient contact is obtained by pumping the mixture of JP-5 and caustic through a coiled pipe.

The JP-5 must be desulfurized in the liquid phase because of its relatively high boiling range. The catalyst is active only in a narrow temperature range between 600°F to 650°F. The JP-5 remains liquid within this temperature range at elevated pressure. The liquid phase system is advantageous because latent heat of vaporization does not have to be provided, however, space velocity considerations favor the gas phase reaction. The space velocity of the liquid phase system is one-fourth of the space velocity of the gas phase reaction.

#### B. HYDROCARBON REFORMING

The processing of the JP-5 to form hydrogen is based on operational experience in naphtha reforming. Minor modifications were provided due to the unique service environment. However, the basic reforming reactions are identical and tests previously conducted in similar development work served to establish process conditions and design criteria. These tests indicated that a steam to carbon ratio of 3 to 1 is the minimum ratio which can be successfully used in continuous operation with this type of feed stock. This steam to carbon ratio is used in all calculations because it is known, by experience, to be a safe value. It should be pointed out that the development program might result in a lower steam requirement.

The first processing step involves the catalytic steam reforming of JP-5 to convert hydrocarbons into hydrogen, CO, CO<sub>2</sub> with only a minor amount of residual methane.

This conversion is accomplished without deposition of elemental carbon mainly by choice of temperature and steam ratio. Operating at temperatures in excess of 1700°F. with an inlet steam to carbon ratio of at least 3-1, the kinetics of all competing reactions involved are sufficiently influenced to make the deposition of carbon impossible.

The 1700°F exit temperature of the reformed gas is low enough to prevent overheating of the catalyst and insure long catalyst life. This life should be in the order of several years. This temperature is high enough to produce a process gas which is low in methane.

The process operating pressure of 250 psig is required in order to provide the necessary differential pressure at the diffuser. Pressures of 250 psig are not unusual in Chemico's reforming practice. Commercial plants reforming natural gas operate at pressures as high as 300 psig.

#### C. CARBON MONOXIDE CONVERSION

The CO-OX "shift" reaction between carbon monoxide and steam to yield additional free hydrogen follows conventional practice. A newer type of catalyst will be employed, which permits operation at a lower temperature level. This is advantageous because the heat of the gas leaving the reformer will be more efficiently utilized and a CO-OX exit gas is produced which contains a lower concentration of carbon monoxide.

#### D. HYDROGEN RECOVERY

Two methods for the final recovery of pure hydrogen from the process gas stream were studied in detail. The method which was selected used a diffuser, while the alternate method involved the selective removal of carbon dioxide from the process gas stream followed by methanation of residual carbon monoxide and carbon dioxide. Thus the alternate method required the recycle of methane from the fuel cell to the reformer. In addition, either a liquid absorbent or a solid absorbent could be used to remove the carbon dioxide. The use of a liquid absorbent would involve serious design problems, due to the necessity of obtaining good gas-liquid contact regardless of any attitude which the submarine may assume in service. Additional problems would arise in handling the large quantities of liquid involved, and in regeneration of the absorbent. Similar considerations applied with respect to the use of a solid adsorbent, since large quantities of adsorbent would be required and frequent regeneration of the adsorbent would be necessary.

Thus, the diffuser approach to the recovery of pure hydrogen was selected. The diffuser unit is a fixed piece of equipment which will function in any attitude of the submarine. The unit does not require any reactivation during normal operation, and provides high purity hydrogen with a minimum of attention. The design efficiency of the unit in terms of hydrogen recovery is 92%, which is comparable to the efficiency of the alternate method involving carbon dioxide removal and methantion. The major disadvantage of the diffusion method is the relatively high cost.

The use of a diffuser to separate pure hydrogen from a mixed gas stream is an established process technique. Diffusers have been employed commercially to produce high purity hydrogen for special applications. Pure hydrogen has been recovered in this manner from reformed gases and also from gas streams produced by ammonia decomposition. These gas streams contain approximately the same concentration of hydrogen. This is an important factor in determining diffuser efficiency.

The operation of a diffuser is based on the selective diffusion or passage of a specific component of a gas mixture through a permeable membrane. A diffuser consisting of palladium-silver alloy tubes will be used in the process. This type of alloy selectively permits only hydrogen molecules to pass through, and is non-permeable to the other gases in the process gas stream. The product hydrogen stream is of ultra high purity, and is subsequently passed to the fuel cell for usage.

The rate of hydrogen passage or diffusion per unit area of diffuser surface is an important design consideration. This rate depends on two factors, the operating temperature and the mean difference in partial pressure of hydrogen across the membrane. It will be evident that the partial pressure of hydrogen in the off gas from the diffuser affects the mean partial pressure difference and thus limits the rate of diffusion. A compromise is required between the percentage of hydrogen recovery from the feed gas and total surface area of the diffuser. A reasonable recovery of 92% of the inlet hydrogen is provided.

#### E. RESIDUAL OFF-GAS TREATMENT

An oxidizer unit is provided in the process, to convert the combustible components in the waste gases from the diffuser and hydrogen from the desulfurization section to carbon dioxide and water by catalytic oxidation. Oxygen is added to the waste gas stream in controlled stages, and controlled oxidation of the combustibles occurs while in contact with a platinum type catalyst. The generated heat is utilized to produce process steam. The final gas stream is disposable and is discharged overboard, after cooling to recover water by condensation.

F. START UP

The hydrogen generator unit will normally start up at dockside. However, the unit can also be started while the submarine is cruising on the surface under diesel power. It is assumed that the unit is at ambient temperature and 0 psig.

The primary objective during start up is to bring all the equipment up to operating temperatures and pressure. When this is done the unit can be adjusted to produce product hydrogen at the desired rate.

Air will be pumped from the CO<sub>2</sub> compressor into the reformer where it will combust with desulfurized JP-5 to produce a hot gas containing essentially CO<sub>2</sub>, N<sub>2</sub> and steam. Excess air and/or water injection will be used to control the temperature in the reformer. The hot gases leaving the reformer will pass through the equipment and will then be discharged overboard.

Hot gases passing through the oxidizer tubes will heat up the water in the shell and generate steam. This steam will be fed to the reformer. At this time oxygen will replace the air and the JP-5 rate will be adjusted so as to produce a small amount of hydrogen gas in the reformer. The hydrogen from the reformer will be combusted in the oxidizer to provide sufficient heat to produce steam at process conditions. The system will be brought to operating pressure. At this time the diffuser will start to emit product hydrogen. Part of the diffuser feed gas will be sent to the desulfurizer unit where it will react with sulfur bearing JP-5 in the O.S.C. vessel to produce H<sub>2</sub>S and desulfurized JP-5. (Before startup, the desulfurizing unit is filled with sulfur free JP-5.)

The final phase of the startup will be to place the system on automatic control at the desired rate of hydrogen product.

It is estimated that the startup procedure described will take approximately two hours.

G. RESPONSE TIME

The time interval required to change from one rate of hydrogen production to another is called response time. The major factors controlling the response time for the hydrogen generator unit are lag time in the control system and steam generation.

Before the JP-5 rate can be increased to produce more hydrogen, the proper amount of steam must first be generated and fed to the reformer. This is necessary so that the steam to carbon ratio in the reformer is maintained at the required value to prevent carbon formation.

To generate the additional steam, part of the diffuser feed gas is by-passed to the oxidizer where it is combusted with oxygen to provide the necessary heat. This procedure results in a temporary decrease in the existing rate of hydrogen production. As soon as the additional steam is generated, the JP-5 and oxygen can be added to the reformer to produce the desired rate of hydrogen.

Table 1 shows estimated response time at various rates. It is assumed that the unit is running at a rate of 10 lb. / hr. hydrogen. Lowering the hydrogen rate will increase the response time.

Because the oxidizer heat load in this design is limited to 540,000 Btu/hr., the largest rate change that can be made in one step is from 10 to 50 lb. /hr. hydrogen.

It will be noted that the estimated response time from 10 to 50 lb. /hr. hydrogen is faster than from 10 to 25 lb. / hr. This is due to a greater gas flow to the oxidizer which will generate the steam at a faster rate.

The response time from 10 to 75 lb. /hr. hydrogen could be improved if the oxidizer is designed for a heat load of 800,000 Btu/hr. and the minimum rate raised to 20 lb. / hr. This would allow the 75 lb. /hr. rate to be reached in one step.

A reduction in the rate of hydrogen production will occur more rapidly than a corresponding increase because the JP-5 flow can be immediately turned down.

The actual lag time in the control system will have to be determined in a prototype unit.

#### RESPONSE TIME

Product H <sub>2</sub> Rate - lb. /hr.		Temporary Reduction in H <sub>2</sub> Rate - lb. /hr.		Estimated Response Time - Secs.
<u>From</u>	<u>To</u>	<u>From</u>	<u>To</u>	
10	25	10	6.3	27
10	50	10	1.0	23
50	75	50	44	18
10	75*			41

\*Two steps required: 10-50 lb. /hr. H<sub>2</sub>  
50-75 lb. /hr. H<sub>2</sub>

H. ENGINEERING ASSUMPTIONS & FUTURE TEST REQUIREMENTS

1. For the purposes of this study J. P. -5 was considered to be equivalent to  $C_{14}H_{30}$  with the following properties:

Boiling Range	400-540°F
Specific Heat	0.71
Specific Gravity	0.8
Heating Value (Net)	18,927 BTU/#
Sulfur Content	0.4% maximum

2. The following items must be verified by pilot plant test:
  - a. The operating conditions and steam to carbon ratio used in this design must be tested. These conditions have been tested on similar fuels and the tests would merely check previous experience.
  - b. The CO oxidation catalyst would be checked to verify vendor claims as to its efficiency at low operating temperatures.
  - c. The diffuser efficiency must be checked with this particular gas stream. The rate of diffusion must also be checked.
  - d. The operating conditions which will provide complete combustion of the oxidizer off gases to carbon dioxide and water must be determined. The ignition temperatures for the various compounds present must be determined to insure controlled combustion.
  - e. Any necessary modification in the normal operating conditions used in the catalytic conversion of the organic sulfur compounds to hydrogen sulfide must be determined. It is essential that a sulfur free, or at least a low ppm sulfur containing fuel, be used in this process. The technique of caustic scrubbing must be checked.
3. The overall system should be tested to determine if any interaction occurs which might affect the process. In this test actual start-up and response times will be determined.
4. Equipment design must be verified for minimum size and maximum performance.

IV. APPENDIX

A. HEAT AND MATERIAL BALANCES

The following tables show the heat and material balances for the process with 76 pounds per hour of product hydrogen. Since a linear relationship exists, the balances for any rate of production may be determined by use of a factor. The material balances are given in mols per hour. The heat balances are given in B. T. U. 's per hour based on the enthalpy of the component in the gaseous state above 77°F. and the heat of combustion values at 77°F. for the combustible materials. The balance is shown between the reactants and products, below the reactants and above the products.

Standard pressure and temperature for gas flows as shown on the flow diagrams are one atmosphere and 32°F.

OVERALL

HEAT AND MATERIAL BALANCE: 76# H<sub>2</sub> PER HOUR

DATUM: TEMP: 77°F STATE: GAS

	TEMP. °F	TOTAL MOLS PER HOUR	MATERIAL BALANCE			HEAT BALANCE			
						BTU/MOL × 10 <sup>-3</sup>			TOTAL
			C	H <sub>2</sub>	O <sub>2</sub>	H <sub>s</sub>	H <sub>c</sub>	H <sub>s</sub> + H <sub>c</sub>	
IN:									
C <sub>14</sub> H <sub>30</sub>	77 °F	1.36	19.0	20.4			3,775	3,775	5,093,000
O <sub>2</sub>	77 °F	10.20			10.20				
H <sub>2</sub> O	140 °F	17.80		17.80	8.90	0.54		0.54	9,700
LESS HEAT REQUIRED TO CONVERT 17.8 H <sub>2</sub> O LIQUID TO GAS									— 300,000
BALANCE			19.0	38.2	19.10	BALANCE			4,802,700
OUT:									
H <sub>2</sub>	140 °F	38.0		38.0		0.45	104.04	104.49	3,980,000
CO <sub>2</sub>	120 °F	19.0	19.0		19.0	0.41		0.41	7,700
H <sub>2</sub> O	120 °F	0.2		0.2	0.1	0.35		0.35	70
HEAT REJECTED TO COOLING LOOP									814,930

THERMAL EFFICIENCY: ASSUME 50,000 BTU \* PROCESS REQ.

$$T.E = \frac{\text{HEAT VALUE H}_2 \times 100}{\text{HEAT VALUE C}_{14}\text{H}_{30} + \text{PROCESS REQ.}} = \frac{3,980,000 \times 100}{5,093,000 + 50,000} = 77.5\%$$

\*BASED ON OPERATION OF ELECTRICAL EQUIPMENT AND INSTRUMENTS BUT NOT COMPRESSOR.

REFORMER 26.1 R-1

HEAT AND MATERIAL BALANCE: 76# H<sub>2</sub> PER HOUR

DATUM: TEMP: 77° F

STATE: GAS

	TEMP. °F	TOTAL MOLS PER HOUR	MATERIAL BALANCE			HEAT BALANCE			
						BTU/MOL × 10 <sup>-3</sup>			TOTAL
			C	H <sub>2</sub>	O <sub>2</sub>	H <sub>s</sub>	H <sub>c</sub>	H <sub>s</sub> + H <sub>c</sub>	
IN:									
C <sub>14</sub> H <sub>30</sub>	500°	1.36	19.0	20.4				3,805.0	5,174,800
H <sub>2</sub> O	1625°	47.10		47.10	23.55	14.060			662,210
O <sub>2</sub>	77°	7.65			7.65				
OUT:			19.0	67.50	31.20	BALANCE			5,837,010
CO <sub>2</sub>	1700°	8.20	8.20		8.20	19.15		19.15	157,030
CO	1700°	10.60	10.60		5.30	12.10	121.74	133.84	1,418,757
CH <sub>4</sub>	1700°	0.20	0.20	0.40		23.00	345.16	368.16	73,652
H <sub>2</sub> O	1700°	35.40		35.40	17.70	14.83		14.83	525,053
H <sub>2</sub>	1700°	31.7		31.70		11.50	104.04	115.54	3,662,513

CO CONVERTER: 26.1 R-4

HEAT AND MATERIAL BALANCE: 76 # H<sub>2</sub> PER HOUR

DATUM: TEMP: 77° F STATE: GAS

	TEMP. ° F	TOTAL MOLS PER HOUR	MATERIAL BALANCE			HEAT BALANCE			
						BTU/#MOL × 10 <sup>-3</sup>			TOTAL
			C	H <sub>2</sub>	O <sub>2</sub>	H <sub>s</sub>	H <sub>c</sub>	H <sub>s</sub> + H <sub>c</sub>	
IN :									
CO <sub>2</sub>	450°	8.20	8.20		8.20	3.60		3.60	29,500
CO	450°	10.60	10.60		5.30	2.60	121.74	124.34	1,320,000
CH <sub>4</sub>	450°	0.20	0.20	0.40		3.60	345.16	348.76	69,800
H <sub>2</sub> O	450°	36.90		36.90	18.45	3.10		3.10	114,200
H <sub>2</sub>	450°	31.70		31.70		2.60	104.04	106.64	3,382,000
OUT :			19.00	69.00	31.95	BALANCE			4,915,500
CO <sub>2</sub>	650°	17.6	17.60		17.60	5.89		5.89	103,800
CO	650°	1.2	1.20		0.60	4.06	121.74	125.80	151,200
CH <sub>4</sub>	650°	0.2	0.20	0.40		6.05	345.16	351.21	70,400
H <sub>2</sub> O	650°	27.50		27.50	13.75	4.90		4.90	134,100
H <sub>2</sub>	650°	41.10		41.10		4.14	104.04	108.18	4,456,000

DIFFUSER 26.2 R-1

HEAT AND MATERIAL BALANCE: 76# H<sub>2</sub> PER HOUR

DATUM: TEMP: 77°F STATE: GAS

	TEMP. °F	TOTAL MOL PER HOUR	MATERIAL BALANCE			HEAT BALANCE			
						BTU/MOL × 10 <sup>-3</sup>			TOTAL
			C	H <sub>2</sub>	O <sub>2</sub>	H <sub>s</sub>	H <sub>c</sub>	H <sub>s</sub> + H <sub>c</sub>	
IN:									
CO <sub>2</sub>	600°	17.6	17.6		17.6	5.47		5.47	96,400
CO	600°	1.2	1.2		0.6	3.71	121.74	125.45	150,400
CH <sub>4</sub>	600°	0.2	0.2	0.4		5.46	345.16	350.62	71,300
H <sub>2</sub> O	600°	2.9		2.9	1.45	4.67		4.67	13,600
H <sub>2</sub>	600°	41.1		41.1		3.66	104.04	107.70	4,440,000
OUT:			19.0	44.4	19.65	BALANCE			4,773,700
CO <sub>2</sub>	600°	17.6	17.6		17.6	5.47		5.47	96,400
CO	600°	1.2	1.2		0.6	3.71	121.74	125.45	150,400
CH <sub>4</sub>	600°	0.2	0.2	0.4		5.46	345.16	350.62	71,300
H <sub>2</sub> O	600°	2.9		2.9	1.45	4.67		4.67	13,600
H <sub>2</sub>	600°	3.1		3.1		3.66	104.04	107.70	336,000
H <sub>2</sub> TO FUEL CELL									
H <sub>2</sub>	600°	38.0		38.0		3.66	104.04	107.70	4,104,000

OXIDIZER: 26.2 H-1

HEAT AND MATERIAL BALANCE: 76 # H<sub>2</sub> PER HOUR

DATUM: TEMP: 77°F STATE: GAS

	TEMP. °F	TOTAL MOLS PER HOUR	MATERIAL BALANCE			HEAT      BALANCE			
						BTU/MOL × 10 <sup>-3</sup>			TOTAL
			C	H <sub>2</sub>	O <sub>2</sub>	H <sub>s</sub>	H <sub>c</sub>	H <sub>s</sub> + H <sub>c</sub>	
IN:									
CO <sub>2</sub>	600°	17.6	17.6		17.6	5.47		5.47	96,400
CO	600°	1.2	1.2		0.6	3.71	121.74	125.45	150,400
CH <sub>4</sub>	600°	0.2	0.2	0.4		5.46	345.16	350.62	71,300
H <sub>2</sub> O	600°	2.9		2.9	1.45	4.67		4.67	13,600
H <sub>2</sub>	600°	3.1		3.1		3.66	104.04	107.70	336,000
O <sub>2</sub>	77°	2.55			2.55				
OUT:			19.0	6.4	22.2	BALANCE			667,700
CO <sub>2</sub>	500°		19.0		19.0	4.23		4.23	80,500
H <sub>2</sub> O	500°			6.4	3.2	3.48		3.48	22,300
LATENT HEAT									
H <sub>2</sub> O	400°F	38.0							564,900

HEAT BALANCE: 76# HYDROGEN PER HOUR  
HEAT EXCHANGER DUTY, HEAT LOAD AND HEAT REJECTION.

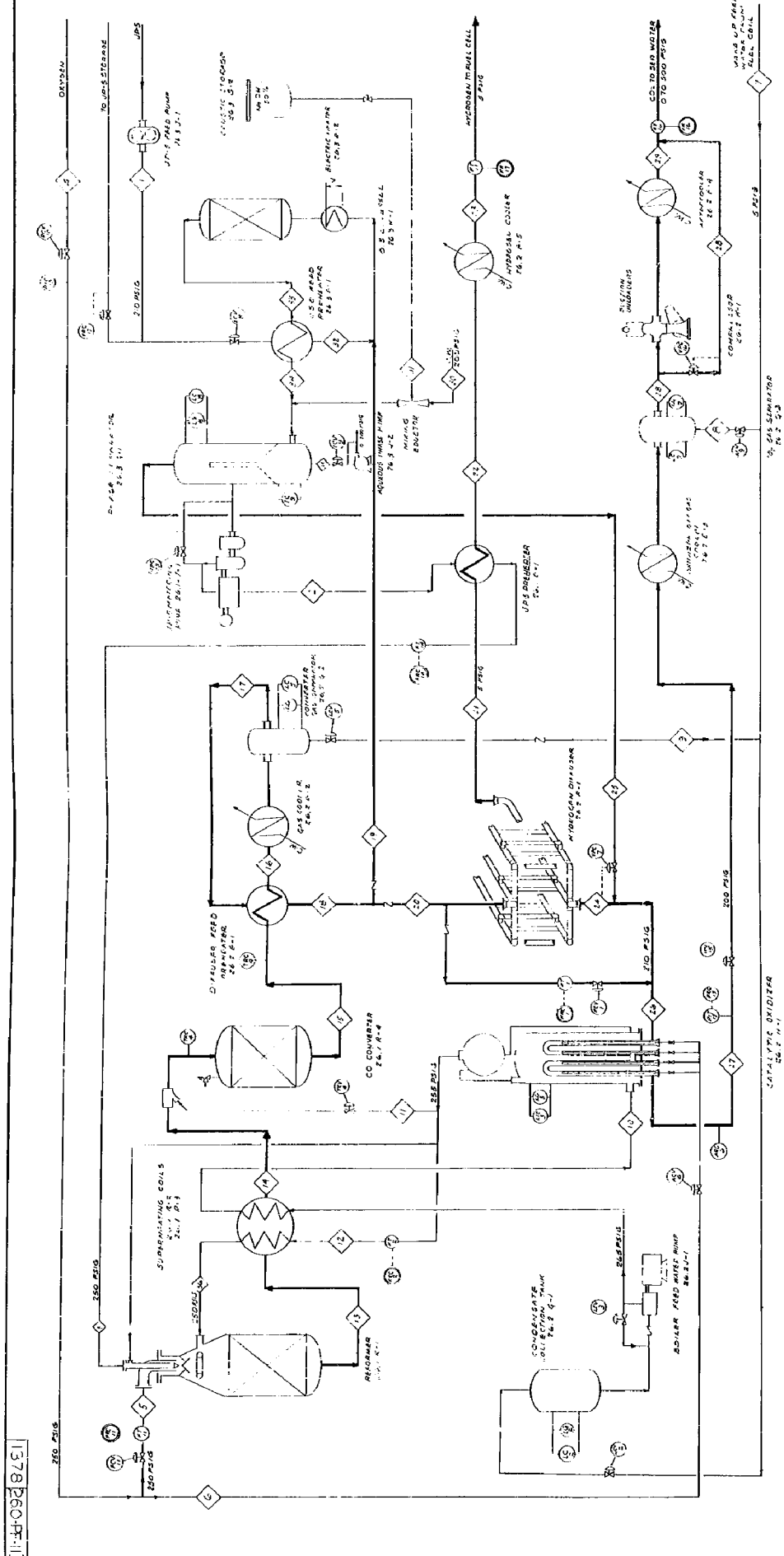
HEAT EXCHANGER NO.	HOT SIDE		HEAT LOAD BTU/HOUR	COLD SIDE		HEAT REJECTION BTU/HOUR
	IN	OUT		IN	OUT	
26.1 R-2	P.G.-2 1700°F	P.G.-2 960°F	561,000	STEAM 400°F	STEAM 1625°F	
26.1 R-3	P.G.-2 960°F	P.G.-2 450°F	360,000	WATER 170°F	WATER & STEAM 400°F	
26.2 E-1	P.G.-4 650°F	P.G.-4 370°F	227,200	P.G.-5 200°F	P.G.-5 600°F	
26.2 E-2	P.G.-4 370°F	P.G.-4 200°F	566,600	COOLING LOOP		566,600
26.1 E-1	HYDROGEN 600°F	HYDROGEN 300°F	81,000	C <sub>14</sub> H <sub>30</sub> 77°F	C <sub>14</sub> H <sub>30</sub> 500°F	
26.2 E-3	P.G.-7 <sup>a</sup> 500°F	P.G.-7 <sup>a</sup> 120°F	204,000	COOLING LOOP		204,000
26.2 E-4	AFTERCOOLER 230°F		19,800	COOLING LOOP		19,800
26.3 E-1	C <sub>14</sub> H <sub>30</sub> 600°F	C <sub>14</sub> H <sub>30</sub> 100°F	95,700	C <sub>14</sub> H <sub>30</sub> 77°F	C <sub>14</sub> H <sub>30</sub> 577°F	
26.3 E-2	ELECTRIC HEATER		4,200	C <sub>14</sub> H <sub>30</sub> 577°F	C <sub>14</sub> H <sub>30</sub> 600°F	
26.2 E-5	HYDROGEN 300°F	HYDROGEN 140°F	44,330	COOLING LOOP		44,330

P.G. = PROCESS GAS

B. FLOW DIAGRAMS

1. Process Flow Diagram (26.0-PF-10)
2. Engineering Flow Diagram (26.0-PF-11)



[illegible]

NOTES:

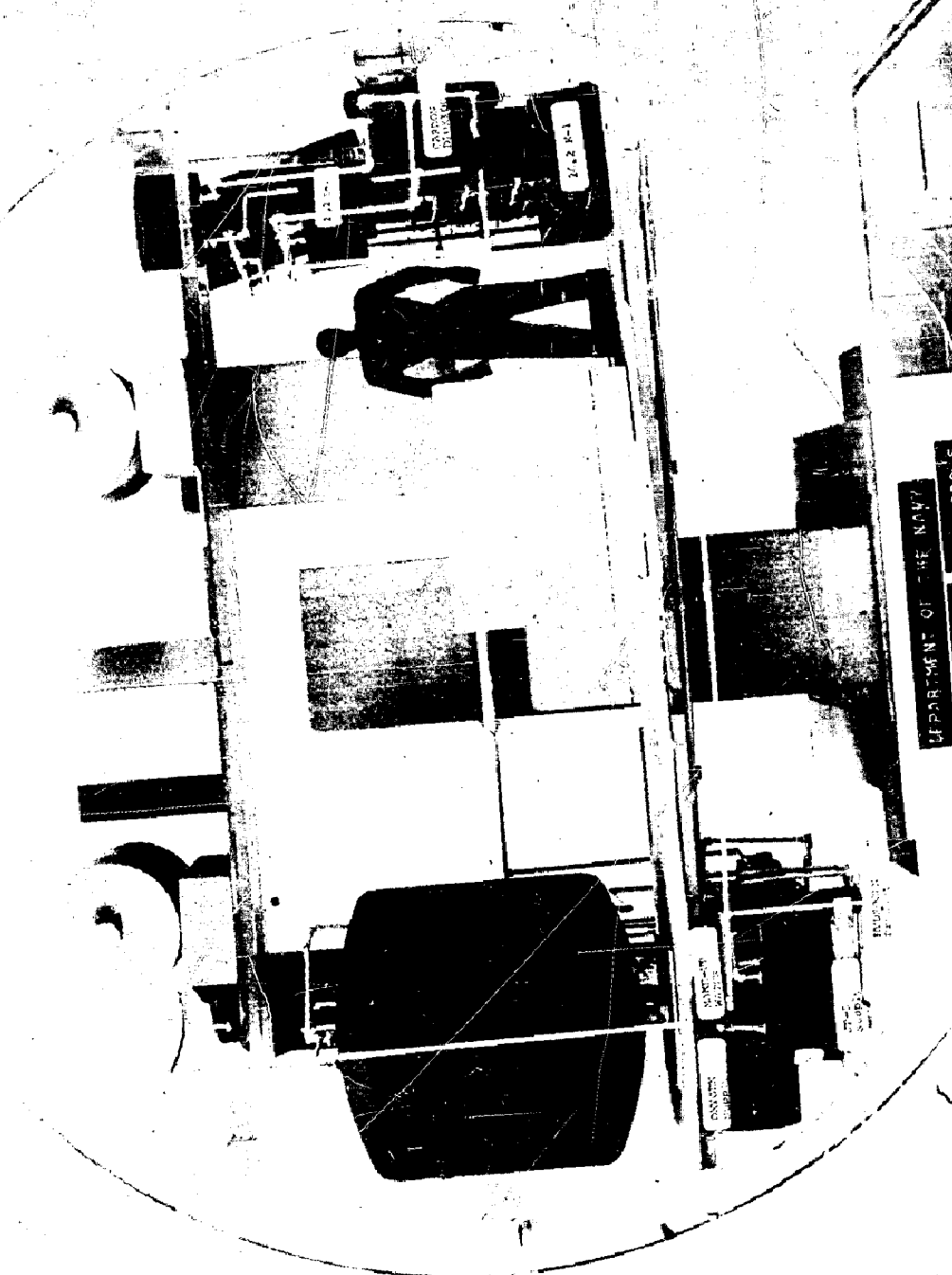
- 1. SYSTEM MEASUREMENTS FACETS WHERE NOTED ARE:  
P4008 - 200 PWS  
P4009 - 200 PWS
- 2. TENDENCY TO BE OVERLY  
INDICATING INSTRUMENTS AND PLANS  
ARE NOT SHOWN
- 3. CMC MEASUREMENTS - 200 PWS - 1.78

[illegible]

C. PHOTOGRAPHS OF MODEL SHOWING  
LAYOUT OF EQUIPMENT

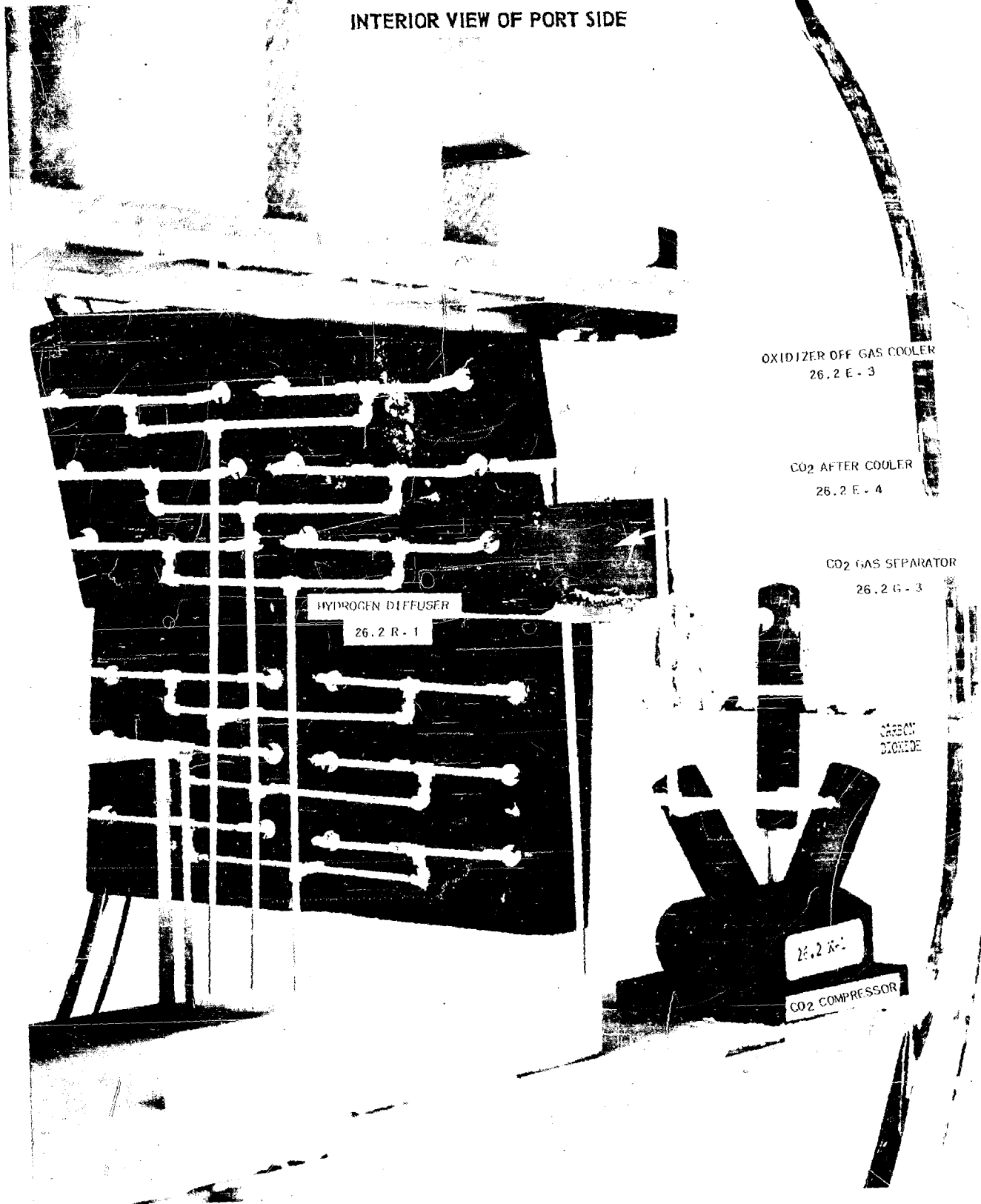
1. Looking Aft
2. Interior View of Port Side
3. Equipment Viewed Through Starboard Side of Hull

LOOKING AFT



DEPARTMENT OF THE NAVY  
BUREAU OF SHIPS CODE 660'S  
HYDROGEN GENERATION PLANT  
CHEMICAL CONSTRUCTION CORPORATION

INTERIOR VIEW OF PORT SIDE



# EQUIPMENT VIEWED THROUGH STARBOARD SIDE OF HULL

O.S.C. (ORGANIC SULFUR CONVERTER)  
26.3 H - 1

CATALYTIC OXIDIZER  
26.2 H - 1

ELECTRIC HEATER  
26.3 E - 2

PHASE SEPARATOR  
26.3 G - 1

INTEGRATED REFORMER

26.1 R - 1	REFORMER
26.1 R - 2	SUPERHEATING COIL
26.1 R - 3	WATER HEATING COIL
26.1 R - 4	CO CONVERTER

CONVERTER GAS SEPARATOR  
26.2 G - 2

CAUSTIC STORAGE  
26.3 G - 2

JP - 5 FEED PUMP  
26.3 J - 1

26.2 E - 1

26.3 E - 2

26.2 J - 1

CONVERTER GAS COOLER  
26.2 E - 2

O.S.C. FEED PREHEATER  
26.3 E - 1

JP - 5 PREHEATER  
26.1 E - 1

DIFFUSER FEED PREHEATER  
26.2 E - 1

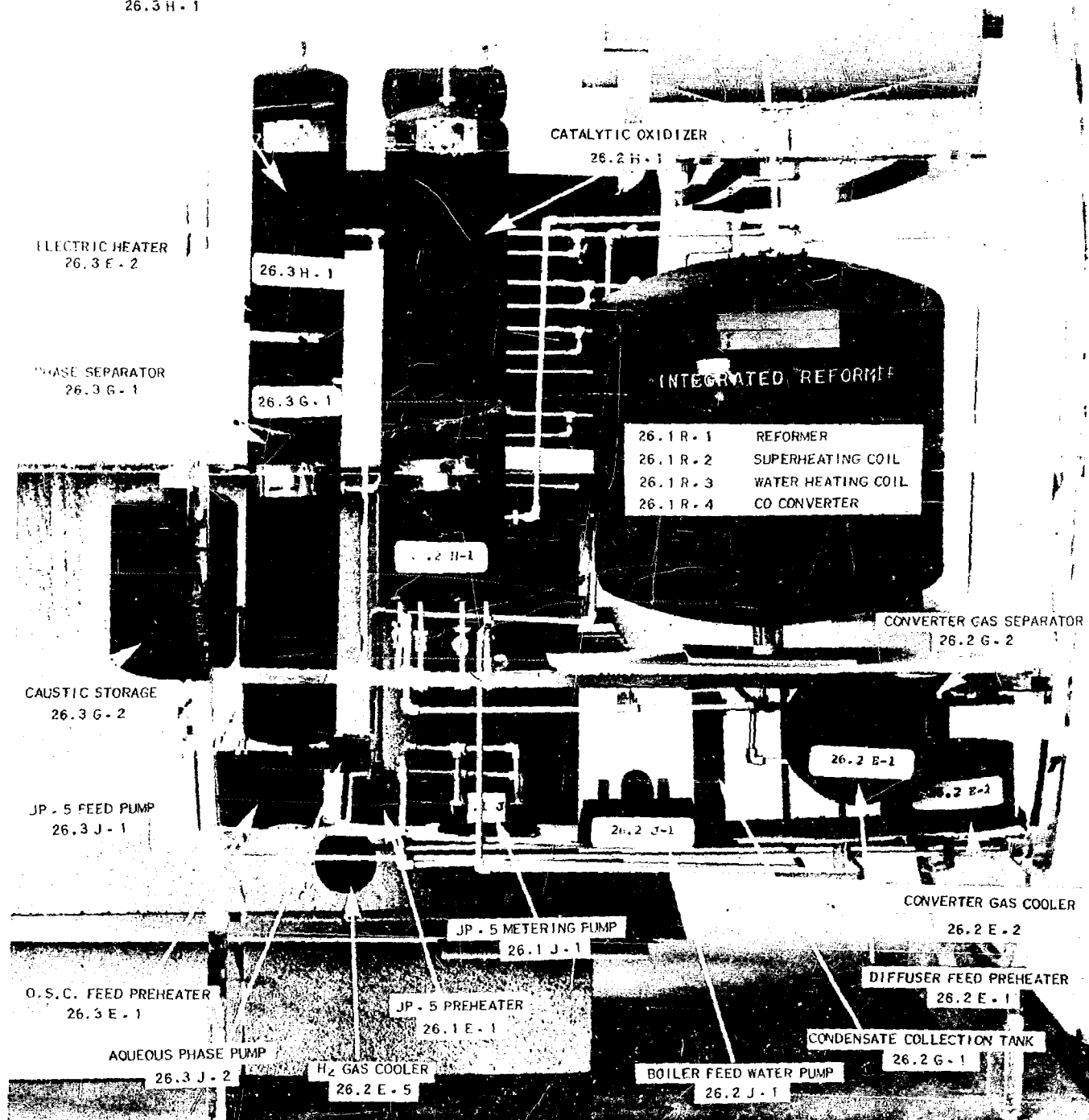
AQUEOUS PHASE PUMP  
26.3 J - 2

H<sub>2</sub> GAS COOLER  
26.2 E - 5

CONDENSATE COLLECTION TANK  
26.2 G - 1

BOILER FEED WATER PUMP  
26.2 J - 1

JP - 5 METERING PUMP  
26.1 J - 1



CHEMICAL CONSTRUCTION CORPORATION

Prepared by: A. Bernas  
A. Chevion  
A. Phillips

Approved by: G. R. James  
W. J. Rosenbloom

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